Design, optimization and techno-economic feasibility study of absorption refrigeration system using ammonia + water as working pair

Submitted by

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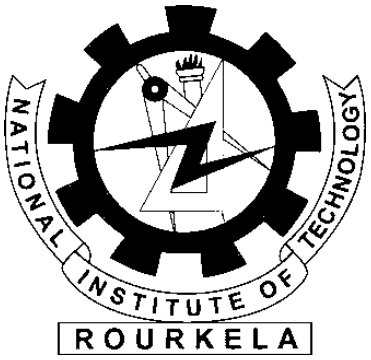
In partial fulfilment of the requirements for the degree in

Bachelor of Technology in Chemical Engineering

Under the esteemed supervision of

Dr. Madhusree Kundu



DEPARTMENT OF CHEMICAL ENGINEERING

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May, 2015

National Institute of Technology Rourkela

CERTIFICATE

This is to certify that the project report entitled, "**Design, optimization and techno-economic feasibility study of absorption refrigeration system using ammonia + water as working pair**." submitted by **Manish Biswal (111CH0267)** in partial fulfilments for the requirements for an award of Bachelor of Technology Degree in Chemical Engineering at National Institute of Technology Rourkela is prepared by him under my supervision and guidance and this work is not been submitted elsewhere for a degree.

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ABSTRACT

Absorption refrigeration systems are the cooling systems that use heat as input for operation instead of electricity. This type of refrigeration system reduces the fossil fuel consumption, indirectly restricts the CO_2 emission to the atmosphere, and restricting at the same time the usage of toxic refrigerants. They generally use refrigerant/absorbent pair for operation. NH_3+ H_2O and LiBr+H_2O are the most common working pair available in the market. In ammonia-water system, the absorbent is water, generally weak solution of ammonia-water and the refrigerant is ammonia. This system consists of an absorber, an evaporator, a condenser, a generator, 2 heat exchangers, 2 valves, a solution pump and an absorber. The dephlegmator and the rectification column are used to decrease the amount of water carried away with ammonia during the regeneration The major problems related to this system is the difference in boiling point of ammonia and water system. This creates a problem in generator section of the system, where both ammonia and water are generated from the solution. But it has a lot of advantages over LiBr system where crystallization occurs, also this has to be operated under nearly vacuum conditions, moreover corrosion problem arise in LiBr system. Ammonia + water as the working pair is taken and the cycle was operated for various temperature ranges of the evaporator, generator and absorber, we also worked on different no. of stages to check the COP of the system. The evaporator temperature has a very insignificant effect on the COP of the system and thus does not change the COP even if we increase the temperature. The generator has a very vivid impact on the COP. As we increase the generator temperature the COP value decreases. Thus further concreting our base for using low grade heat as the source of power for operating the generator. Further no. of stages of the separator also effects the COP of the system. Increasing the no. of stages increases the COP up to a certain level but on further increasing the stages the COP becomes constant. Thus the optimum conditions comes as- T_G = 70 C, T_E = 5 C, T_C = 33 C, T_A = 30 C and the optimum no. of stages for the separator comes to 8.The feasibility study also suggests that the use same capacity (1.5TR) ARS instead of VCR is profitable by an amount of INR 8K annually. From the feasibility analysis we observe that for a refrigeration of 1 TR, we require 23 MWh low grade heat at a temperature of 70ºC. Further, if we change the refrigeration load the requirement of low grade heat also changes.

Keywords- ARS, VCR, COP, low grade heat, optimization, ammonia, water

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NOMENCLATURE

Symbol	Meaning
Q_C	Condenser heat duty (in kW)
Q_E	Evaporator heat duty (in kW)
Q_A	Absorber heat duty (in kW)
Q_D	Dephlegmator heat duty (in kW)
W_P	Pump work (in kW)
m	mass flowrate of pure refrigerant (in kg/s)
m_{ss}	mass flowrate of strong solution (in kg/s)
m_{ws}	mass flowrate of weak solution (in kg/s)
λ	circulation ratio
COP	Coefficient of performance
T	Temperature (in ºC)
P	Pressure (in bar)
T_E	Temperature of the evaporator (in ºC)
T_A	Temperature of the absorber (in ºC)
T_G	Temperature of the generator (in ºC)
T_C	Temperature of the condenser (in ºC)
HE1	Sub cooling heat exchanger
HE2	Solution heat exchanger
TR	Tonnes of refrigeration
$	US Dollars
Rs	Indian Rupee
VCR	Vapour compression refrigeration system
ARS	Absorption refrigeration system

CHAPTER- 1

INTRODUCTION

1.1 INTRODUCTION-

Mechanical courses of action far and wide utilize a great deal of heat vitality by blazing fossil energizes to create steam or essentially heat. A few spots like the Data centres, lodgings, shopping centres and multi-storeyed structures create a great deal of heat, as expanding the temperature of the whole place. The early advancement of a absorption cycle goes back to the 1700's. It was realized that ice could be delivered by a vanishing of immaculate water from a vessel contained inside an emptied holder in the vicinity of sulfuric corrosive. In 1810, ice could be produced using water in a vessel, which was joined with another vessel containing sulfuric corrosive. As the corrosive absorbed water vapour, bringing on a lessening of temperature, layers of ice were shaped on the water surface. The real issues of this framework were erosion and spillage of air into the vacuum vessel. This machine took out a US patent in 1860. Machines in light of this patent were utilized to make ice and store nourishment. It was utilized as a fundamental plan as a part of the early period of refrigeration advancement [1].

We generally use the centralised air conditioning or the air coolers to remove the heat from this places. But the hot air from these coolers are thrown out into the ambience, this subsequently increases the overall temperature of the globe. This heat can be used for heat-operated absorption refrigeration systems. This greatly reduces the global problems like **CO_2** emission and the depletion of **ozone layer** due to release of **CFCs** into the environment. So there is an imminent need for solving these issues. Utilising this wasted heat for the operation of vapour **absorption refrigeration system (ARS)** greatly solves this issues. An absorption refrigeration system utilises a working pair that also lowers the **CFC** refrigerants used in the vapour compression refrigeration system. Vapour absorption refrigeration system utilises a compressor, a condenser, an expansion valve, an evaporator and a refrigerant as the working fluid. Whereas an absorption refrigeration system uses a chemical compressor as an exception from the conventional system. A **chemical compressor** uses an **absorber** and a **generator** (separator). The refrigerant coming out of the evaporator is absorbed in an absorbent, generally a weak solution of the refrigerant, and the absorbed fluid is separated in the generator using heat. The Data Centres have huge servers, for example, the servers of **GOOGLE** or **Facebook** servers produce lot of heat and this heat is removed using air conditioners and air coolers. The heat from this coolers are generally thrown out into the ambience. Further the heat is also rejected in the power generation plants. All this heat are basically low grade heat, but are in the reusable form. Thus an ARS can use the heat rejected from these systems as it operates at a lower temperature. Hence it not only reduces the emission of CO_2 and CFCs into the environment but also utilises the heat rejected from the power generation plants and data centres and malls.

A Vapour Absorption Refrigeration System is similar to a **Vapour Compression Refrigeration** (VCR) System. In both systems the required refrigeration is provided by refrigerants vaporizing in the evaporator. However, in the ARS, a **physico-chemical process** replaces the **mechanical process** of the VCR system and heat rather than a mechanical and electrical energy is used. In the absorption refrigeration system the vapour is drawn from the evaporator by absorption into liquid having high affinity for the refrigerant. The refrigerant is expelled from the solution by application of the heat and its temperature is also increased. This refrigerant in the vapour form passes to the condenser where heat is rejected and refrigerant gets liquefied. This liquid again flows to the evaporator at the reduced pressure and cycle is completed. The elimination of the necessary shaft work has been the prime reason for the economic success of vapour absorption system [3].

ARS has different working pair models, **$LiBr+H_2O$** system, **NH_3+H_2O** system and also ionic liquid system as working pair for refrigeration. In NH_3-H_2O system, water is the absorbent and ammonia is the refrigerant. Whereas in LiBr-H_2O system. LiBr is the absorbent and water is the refrigerant [2].

Apart from these two working pairs, we have ionic liquids as the fluid for refrigeration in modern ARS. Ionic liquid is a very new concept and much work has not been done for this fluid. But this thesis deals with the NH_3+H_2O as the working fluid. The design of a vapour absorption refrigeration system using ammonia-water as working fluid. In this study, **ionic-liquids** (ILs), which are salts in a liquid state usually with organic cations and inorganic anions, are used as an absorbent fluid in a miniature absorption refrigeration system designed for current electronic cooling requirements (i.e., benchmark 100 W/cm2 power dissipation and 85 _C chip temperature). ILs have the character of molten salts, which are moisture and air stable at room temperature. Most ILs are thermally stable to temperatures well above those in vapour compression refrigeration systems, >400 K [4].

1.2 LITERATURE REVIEW-

This thesis deals with the idea of using a vapour absorption refrigeration system for the cooling purposes. But the prime question that arises is what the need of this type of system? As we have conventional VCR systems. This question comes down to listing of the working pros and cons of a VCR and ARS.

Vapour compression refrigeration cycle uses a working fluid, a refrigerant that is circulated through the cycle. The refrigerant is heated in the refrigerated space in the evaporator. The hot refrigerant is passed through a compressor which increases the pressure of the refrigerant and vaporises it. The refrigerant is send to the condenser where it is cooled and condensed at constant pressure, the condensed liquid is sent to evaporator, through the expansion valve, again for taking up heat from the refrigerated space. This kind of system depends on the use of electricity to meet the energy demand. The compressor acts as a source of removal of heat from the refrigerated space [5].

In an absorption refrigeration system a working pair is used, one is the absorbent and the other is the refrigerant. Lithium bromide + water and ammonia + water are the most commonly used working pairs. The absorption refrigeration system basically uses a chemical compressor for the cooling. Here the absorber absorbs the refrigerant and the generator separates the pair. In the process cooling effect is produced. The basic difference in power is the use of thermal power in the ARS. The generator uses the heat as source for operation instead of electricity used by VCR [6].

The working liquid in an absorption refrigeration framework is a parallel arrangement comprising of refrigerant and absorbent. Two cleared vessels are for the most part associated to one another. One vessel contains fluid refrigerant while the other vessel contains a paired arrangement of absorbent/refrigerant. The arrangement in the second vessel will retain refrigerant vapour from the first vessel bringing about pressure to diminish. While the refrigerant vapour is being retained, the temperature of the staying refrigerant will lessen as an after effect of its vaporization. This causes a refrigeration impact to happen inside the main vessel. In the meantime, arrangement inside the second vessel gets to be more weaken due to the higher substance of refrigerant consumed. This is known as the "absorption process". Regularly, the ingestion methodology is an exothermic procedure, in this way, it must reject warmth out to the encompassing with a specific end goal to keep up its absorption

capacity. At whatever point the arrangement can't proceed with the retention process in light of immersion of the refrigerant, the refrigerant must be differentiated out from the weakened arrangement. Heat is typically the key for this detachment process. It is connected to the right vessel to dry the refrigerant from the arrangement. The refrigerant vapour will be consolidated by exchanging warmth to the surroundings. With these methods, the refrigeration impact can be created by utilizing heat vitality. Then again, the cooling impact can't be created ceaselessly as the methodology can't be done at the same time. Accordingly, a retention refrigeration cycle is a mix of these two methods. As the partition methodology happens at a higher weight than the ingestion handle, a dissemination pump is obliged to flow the arrangement. Coefficient of Performance of an absorption refrigeration framework. The work information for the pump is unimportant with respect to the heat data at the generator, along these lines, the pump work is regularly dismissed for the reasons of examination [1].

1.3 ABSORPTION REFRIGERATION SYSTEMS-

Absorption refrigeration systems are the cooling systems that use heat as input for operation instead of electricity. They generally use refrigerant/absorbent pair for operation. NH_3+ H_2O and LiBr+H_2O are the most common working pair available in the market.

Basic Working Principle-

- The ARS consists of a binary solution- refrigerant and absorbent.

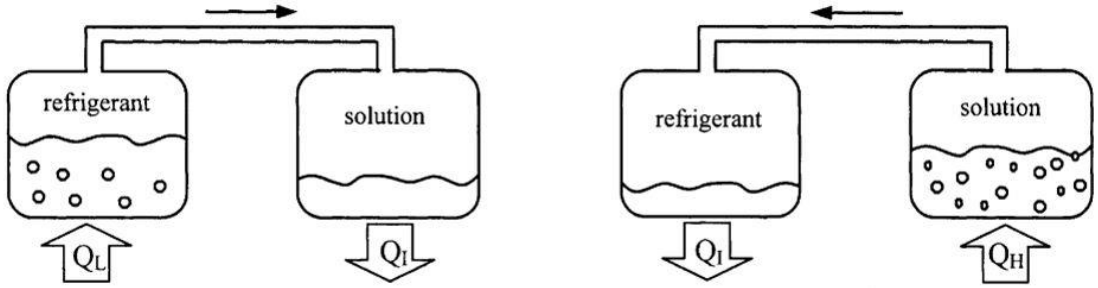


Fig 1.1 Working of a general ARS [1] -

- Two evacuated vessels are generally connected to each other (Fig 1.1).
- In vessel 1, the refrigerant present absorbs the heat from the ambient (which has to be cooled).
- This heats the refrigerant and vaporises it. The vaporisation causes a cooling effect on the refrigerated space.
- The absorbent absorbs the vaporised refrigerant. Since, it is an exothermic process. It releases heat to the ambience.
- The absorption continues till the saturation is reached. Beyond this point no further absorption takes place [16].
- This is where regeneration process is implemented this is basically a separation technique to separate out the refrigerant from the absorbent.
- The heat required for this process is taken from a source. The heating separates the refrigerant and the refrigerant vapours are transferred back to the evaporator.

- These are the two important processes generally used in an absorption refrigeration system. But these two process cannot occur simultaneously. Thus, a cycle for this system is used [9].
- Further, the separation processes occurs at an elevated pressure. Because of this a pump is being used to pressurise the fluid (refrigerant) to the generator.
- Thus heat balance to the system - $Q_E + Q_G + W_P = Q_A + Q_D$
- COP of the system is calculated as- $Q_E/ (Q_G + W_P)$ [14]
- Working fluid for the refrigerant is chosen such that it enhances the overall performance of the system, must be chemically stable, non-toxic and non-explosive.
- In liquid phase they must have miscibility with the operating temperature range,.
- The elevation of boiling point should be as high as possible.
- The refrigerant should have high heat of vaporization and high concentration in the absorbent to maintain low circulation rate between generator and absorber. Refrigerant and absorbent should be of low cost. Transport properties like viscosity, thermal conductivity and diffusion coefficient should be favourable in working temperature and pressure ranges [10].

1.4 OBJECTIVE AND SCOPE-

OBJECTIVE-

- Recovery of low grade heat using absorption refrigeration system.

SCOPE-

- Simulation, design and optimization of absorption refrigeration cycle using ammonia-water as working pair.
- Techno-economic feasibility of absorption refrigeration cycle.

1.5 CHAPTER LAYOUT-

- Chapter 1 deals with introduction to absorption refrigeration system using ammonia-water as working pair
- Chapter 2 deals with the working of ammonia-water absorption refrigeration system.
- Chapter 3 deals with the simulation of absorption refrigeration system using ammonia-water as working pair.
- Chapter 4 deals with Optimization and Design of the ARS.
- Chapter 5 discusses the conclusion and future recommendation.

CHAPTER 2

ABSORPTION REFRIGERATION CYCLE USING NH_3+H_2O AS WORKING PAIR

2.1 AMMONIA+WATER BASED ABSORPTION REFRIGERATION CYCLE-

- In ammonia-water system, the absorbent is water, generally weak solution of ammonia-water and the refrigerant is ammonia.
- The major problems related to this system is the difference in boiling point of ammonia and water system. This creates a problem in generator section of the system, where both ammonia and water are generated from the solution [11].
- Further this system is incompatible with copper and brass and thus, steel is used generally as MoC for the system.
- Also ammonia-water system is flammable so safety issues arise for this system, thus, has to be operated under vigilant conditions.
- But it has a lot of advantages over LiBr system where crystallization occurs also this has to be operated under nearly vacuum conditions, moreover corrosion problem arise in LiBr system.
- This system consists of an absorber, an evaporator, a condenser, a generator, 2 heat exchangers, 2 valves, a solution pump and an absorber.
- The dephlegmator and the rectification column are used to decrease the amount of water carried away with ammonia during the regeneration [6].

The stages in the cycle can be seen from Fig 2.1.

WORKING PRINCIPLE-

- Low temperature and low pressure vapour leaves the evaporator after gaining heat from the refrigerated space and exchanges heat in HE1 with the refrigerant ammonia coming from the condenser.
- The pure ammonia now enters the absorber at state1. The ammonia gets absorbed in a weak solution of ammonia-water sprayed on the absorber which comes from the generator.
- A strong solution of ammonia absorbed in water leaves the absorber at stage 2.
- The stage 2 strong solution is then pumped to the generator pressure using a solution pump. The pressurised strong solution then goes to the heat exchanger where it is preheated by the vapour coming from the generator bottoms [12].
- The heated strong solution then goes to the generator where it is heated for the separation. Here the ammonia vaporise and separates from the water and comes out as ammonia vapours and water comes down through the stripping section of the separator. The ammonia going out contains about 5% of water in it which is then separated using a dephlegmator, this is basically a partial condenser.
- The vapour leaving the rectifier goes to the condenser where it is condensed to liquid ammonia at constant pressure [7].
- The condenser pressure is kept at the high pressure of the system.
- The pure ammonia from the condenser is sent to the sub-cooling heat exchanger where it rejects the heat to the heated ammonia coming from the evaporated.
- The stream from the sub-cooling heat exchanger is sent through a valve to the evaporator.
- The expansion valve releases the pressure from the high pressure stream and sends the refrigerant, again to the evaporator for further cooling.

- The stream coming from the generator bottoms passes through the solution heat exchangers where it rejects its heat to the stream coming from the absorber. In this way, the absorber stream gets pre-heated and the cooled weak solution stream passes through the valve 2.
- The valve releases the pressure from the stream and sends the low pressure stream back to the absorber for absorption of pure ammonia [13].

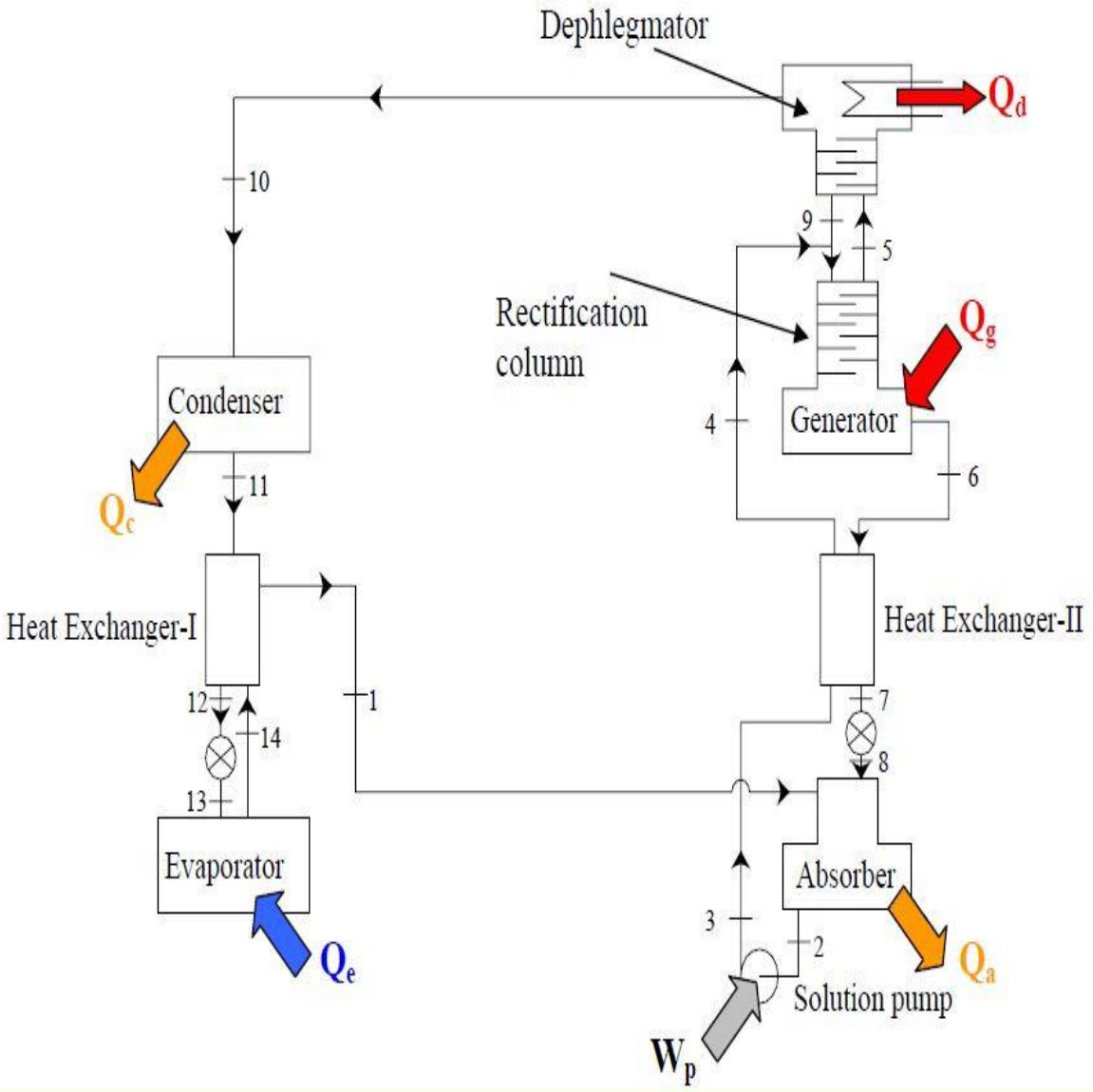


Fig- 2.1 Schematic of ARS using ammonia-water as working pair [2].

Table 2.1-Enthalpy concentration data.

State point	Pressure (bar)	Temperature (ºC)	Concentration(x) (kg of NH_3/kg of H_2O)	Enthalpy (kJ/kg)
1	2	13.9	0.996	
2	2	26.1	0.4	-58.2
3	13.61	26.1	0.4	-56.8
4	13.61	93.3	0.4	253.6
6	13.61	115.6	0.3	369.9
7	13.61	36.1	0.3	
8	2	36.1	0.3	
10	13.61	54.4	0.996	1512.1
11	13.61	36.1	0.996	344.3
12	13.61	30.1	0.996	318.7
13	2	-17.8	0.996	
14	2	4.4	0.996	1442.3

2.2 A SAMPLE CASE STUDY

MASS BALANCE [3]-

For condenser- $\mathbf{m_{10} = m_{11} = m}$

$$\mathbf{Q_C = m\,(h_{10} - h_{11})} \quad (2.1)$$

For evaporator- $\mathbf{m_{13} = m_{14} = m}$

$$\mathbf{Q_E = m\,(h_{13} - h_{14})} \quad (2.2)$$

For absorber- $\mathbf{m_{ss} = m_{ws} + m}$ (2.3)

Circulation ratio- $\boldsymbol{\lambda = m_{ws} / m}$

$$\mathbf{Q_A = m * h_1 + m_{ws} * h_8 - m_{ss} * h_2} \quad (2.4)$$

For valve- enthalpy remains constant. (2.5)

For generator- $\mathbf{Q_G = Q_D + m * \{(h_{10} - h_4) - \lambda\,(h_6 - h_4)\}}$ (2.6)

Pump work- $\mathbf{W_P = m_{ss} * (h_3 - h_2)}$ (2.7)

COP CALCULATION-

For 100 TR refrigeration.

Q_E = **351.7 kW**

Q_D = **88 kW** available

From Table- 2.1

Flowrate- m = **.313 kg/s**

λ = **5.345**, m_{ws} = **1.673 kg/s** and m_{ss} = **1.986 kg/s**

Q_A = **577.4 kW**

Q_C = **365.5 kW**

Q_G = **676.48 kW**

$W_P = \mathbf{2.78\ kW}$

COP Calculation- $\mathbf{COP} = \frac{\textbf{cooling capacity of the system}}{\textbf{heat input by generator+pump work}} = \mathbf{Q_E/(Q_G + W_P)}$ (2.8)

COP = **0.518**

Thus COP calculated is **0.518.**

CHAPTER 3

SIMULATION OF ABSORPTION REFRIGERATION CYCLE USING NH_3+H_2O AS WORKING PAIR

Simulation of ARS using (NH_3 + Water) as working pair has been taken up and presented in this chapter. Following is the sequential approach, which has been adapted for simulating the ARS. Successful simulation may lead to an optimized design, which is energy efficient. Simulations were carried out by varying evaporator temperature, generator temperature and number of stages in the separator, one at a time, by keeping the others constant.

3.1 SIMULATION SOFTWARE-

ASPEN PLUS v8.4

3.2 PROPERTY METHOD-

Peng-Robinson property method

3.3 COMPONENTS-

Ammonia (NH_3)

Water (H_2O)

3.4 BINARY ANALYSIS-

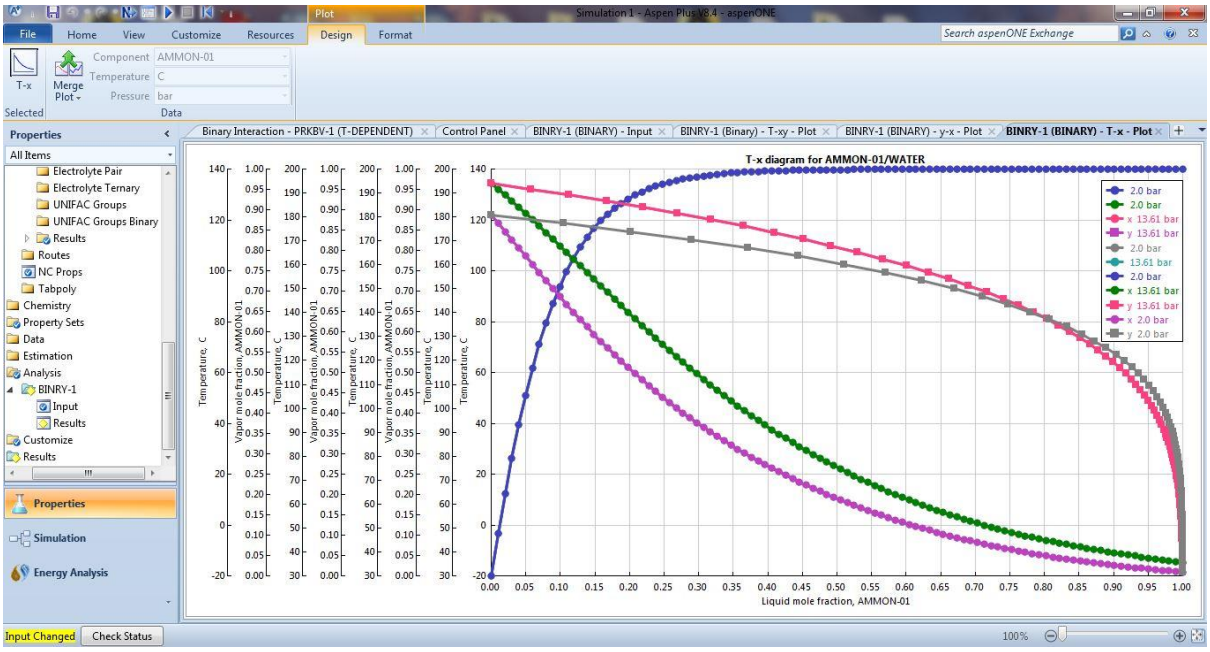


Fig 3.1- T-xy data for pressure at 2 bar and 13.61 bar.

3.5 FLOW SHEET FOR ARS USING AMMONIA-WATER AS WORKING PAIR-

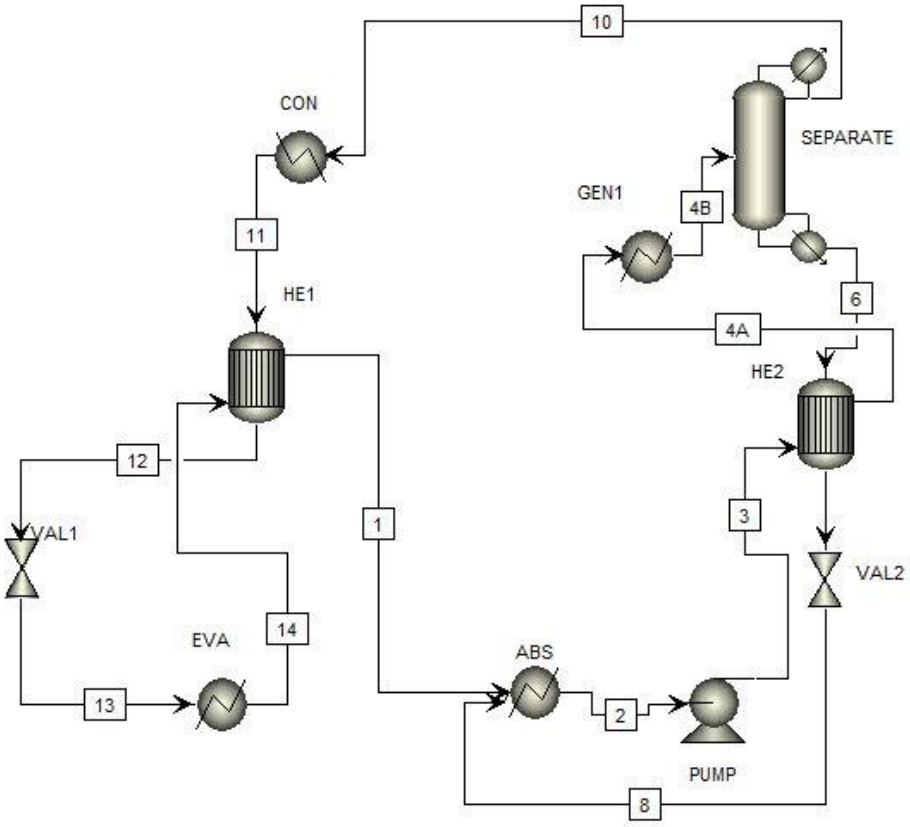


Fig 3.2- Flow sheet showing ARS with ammonia-water as working pair- 1. EVA- Evaporator, 2. ABS- Absorber, 3. Pump, 4. HE2- Heat exchanger 2, 5. Gen1- Generator supplying necessary heat, 6. CON- Condenser, 7. HE1- Heat exchanger 1, 8. VAL1, VAL2- valve 1 and 2 respectively.

The method used for carrying out the component analysis is Peng-Robinson method, this method is found suitable for simulating the NH_3-H_2O system. The simulation is carried out according to the flow sheet in Fig- 3.2.

Pure ammonia from the evaporator takes the heat from the refrigerated space and the stream 14 takes the heated refrigerant through the heat exchanger 1 and sends the refrigerant to the absorber where it is absorbed in the **weak solution** of ammonia and water. The absorbed **strong solution** at elevated temperature is pumped to the generator where the refrigerant is separated from the absorbent using heat. DSTWU column is used for the separation. The pure ammonia from the top is sent to the condenser, a heat exchanger, where the vapour ammonia is condensed at constant temperature and sent to the heat exchanger 1 where it gives out its heat to the refrigerant coming out of the evaporator. The hot stream outlet of the heat exchanger 1 is sent through an expansion valve which releases the pressure of the refrigerant and sends the liquid refrigerant to the evaporator [8].Table 3.1 presents the detail of the connecting streams of the flow-sheet constructed.

Table- 3.1

Stream No.	From	To
1	HE1	Absorber
2	Absorber	Pump
3	Pump	HE2
4A	HE2	Gen1
4B	Gen1	Separator
6	Separator	HE2
7	HE2	Valve 2
8	Valve 2	Absorber
10	Separator	Condenser
11	Condenser	HE1
12	HE1	Valve 1
13	Valve 1	Evaporator
14	Evaporator	HE1

3.6 BASIS-

The basis chosen for simulation is 100 TR refrigeration at evaporator.

Flowrate of pure refrigerant [2],

- $m = Q_E / (h_{14} - h_{13})$
- $m = 351.7 / (1442.3 - 318.7)$ kg/s

  0.313 kg/s

Circulation ratio-

- $\lambda = m_{ws} / m = (x_{10} - x_4) / (x_4 - x_7)$

  $= 5.345$
- $m_{ws} = \lambda * m$

  $=$ **1.673 kg/s**
- $m_{ss} = (1 + \lambda) * m$

  $=$ **1.986 kg/s**

Evaporator temperature [3] = **(5- 10) °C**

Generator temperature = **(70- 90) °C**

Absorber temperature = **(30- 40) °C**

Valve discharge = **2 bar**

Pump discharge = **13.61 bar**

Heat Exchanger efficiency = **1**

Table- 3.2 Stream Result for all streams-

Ammonia-Water refrigeration system														
Stream ID		1	2	3	4A	4B	6	7	8	10	11	12	13	14
From		HE1	ABS	PUMP	HE2	GEN1	SEPARATE	HE2	VAL2	SEPARATE	CON	HE1	VAL1	EVA
To		ABS	PUMP	HE2	GEN1	SEPARATE	HE2	VAL2	ABS	CON	HE1	VAL1	EVA	HE1
Phase		VAPOR	LIQUID	LIQUID	LIQUID	VAPOR	LIQUID	LIQUID	MIXED	VAPOR	LIQUID	LIQUID	MIXED	VAPOR
Substream: MIXED														
Mole Flow	kmol/sec													
AMMON-01		.0182319	.0183052	.0183052	.0183052	.0183052	7.32209E-5	7.32209E-5	7.32209E-5	.0182319	.0182319	.0182319	.0182319	.0182319
WATER		2.77986E-7	6.94966E-5	6.94966E-5	6.94966E-5	6.94966E-5	6.92186E-5	6.92186E-5	6.92186E-5	2.77986E-7	2.77986E-7	2.77986E-7	2.77986E-7	2.77986E-7
Total Flow	kmol/sec	.0182322	.0183747	.0183747	.0183747	.0183747	1.42439E-4	1.42439E-4	1.42439E-4	.0182322	.0182322	.0182322	.0182322	.0182322
Total Flow	kg/sec	.3105060	.3130000	.3130000	.3130000	.3130000	2.49398E-3	2.49398E-3	2.49398E-3	.3105060	.3105060	.3105060	.3105060	.3105060
Total Flow	cum/sec	.2268248	5.25865E-4	5.27583E-4	5.27929E-4	.0354278	3.48038E-6	3.28501E-6	1.47887E-4	.0306062	5.27018E-4	5.13500E-4	.0277854	.2081172
Temperature	K	304.0719	303.1500	304.3887	304.6361	343.1500	346.3750	313.1500	287.2866	309.0154	306.1500	296.1500	254.6188	280.1500
Pressure	N/sqm	2.00000E+5	2.00000E+5	1.36100E+6	1.36100E+6	1.36100E+6	1.36100E+6	1.36100E+6	2.00000E+5	1.36100E+6	1.36100E+6	1.36100E+6	2.00000E+5	2.00000E+5
Vapor Frac		1.000000	0.0	0.0	0.0	1.000000	0.0	0.0	.0868882	1.000000	0.0	0.0	.1456654	1.000000
Liquid Frac		0.0	1.000000	1.000000	1.000000	0.0	1.000000	1.000000	.9131118	0.0	1.000000	1.000000	.8543346	0.0
Solid Frac		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Enthalpy	J/kmol	-4.5801E+7	-6.7524E+7	-6.7412E+7	-6.7390E+7	-4.5679E+7	-1.7324E+8	-1.7601E+8	-1.7601E+8	-4.6317E+7	-6.6378E+7	-6.7240E+7	-6.7240E+7	-4.6662E+7
Enthalpy	J/kg	-2.6894E+6	-3.9640E+6	-3.9574E+6	-3.9562E+6	-2.6816E+6	-9.8942E+6	-1.0053E+7	-1.0053E+7	-2.7196E+6	-3.8976E+6	-3.9482E+6	-3.9482E+6	-2.7399E+6
Enthalpy	Watt	-8.3506E+5	-1.2407E+6	-1.2387E+6	-1.2383E+6	-8.3933E+5	-24675.89	-25070.84	-25070.84	-8.4446E+5	-1.2102E+6	-1.2259E+6	-1.2259E+6	-8.5076E+5
Entropy	J/kmol-K	-1.0412E+5	-1.8656E+5	-1.8632E+5	-1.8625E+5	-1.1634E+5	-1.6638E+5	-1.7480E+5	-1.7431E+5	-1.2097E+5	-1.8597E+5	-1.8883E+5	-1.8772E+5	-1.0707E+5
Entropy	J/kg-K	-6113.841	-10952.29	-10937.79	-10933.64	-6829.475	-9502.684	-9983.195	-9955.568	-7103.395	-10919.75	-11087.66	-11022.44	-6287.027
Density	kmol/cum	.0803804	34.94184	34.82808	34.80530	.5186515	40.92640	43.36040	.9631624	.5957049	34.59513	35.50588	.6561795	.0876057
Density	kg/cum	1.368924	595.2093	593.2715	592.8833	8.834857	716.5838	759.2010	16.86409	10.14520	589.1750	604.6855	11.17512	1.491977
Average MW		17.03057	17.03428	17.03428	17.03428	17.03428	17.50909	17.50909	17.50909	17.03057	17.03057	17.03057	17.03057	17.03057
Liq Vol 60F	cum/sec	9.76470E-4	9.81641E-4	9.81641E-4	9.81641E-4	9.81641E-4	5.17094E-6	5.17094E-6	5.17094E-6	9.76470E-4	9.76470E-4	9.76470E-4	9.76470E-4	9.76470E-4

Stream results for the above simulation is tabulated in Table 3.2

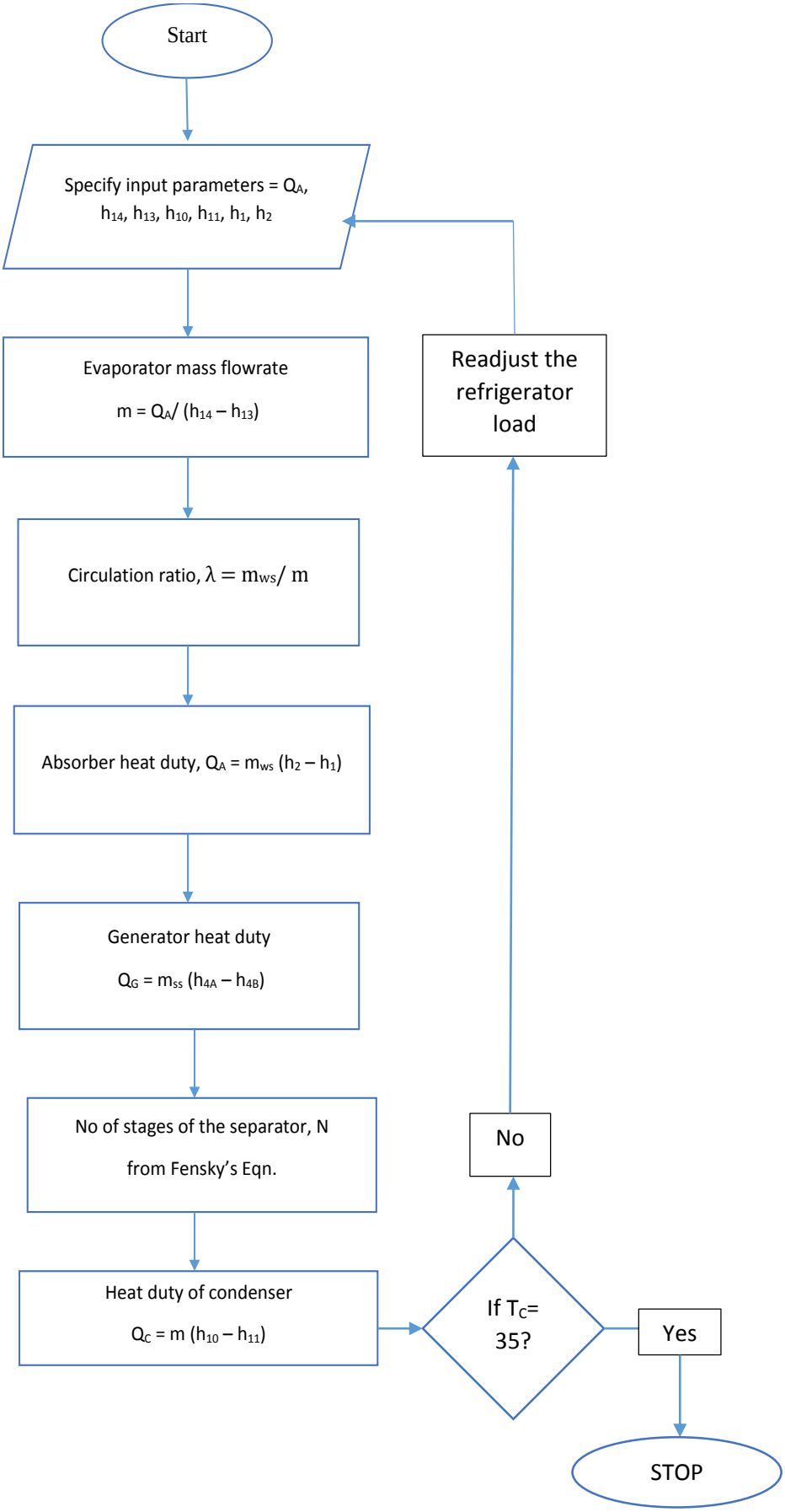


Fig- 3.3- The overall simulation flow sheet

3.7 INPUTS TO VARIOUS BLOCKS

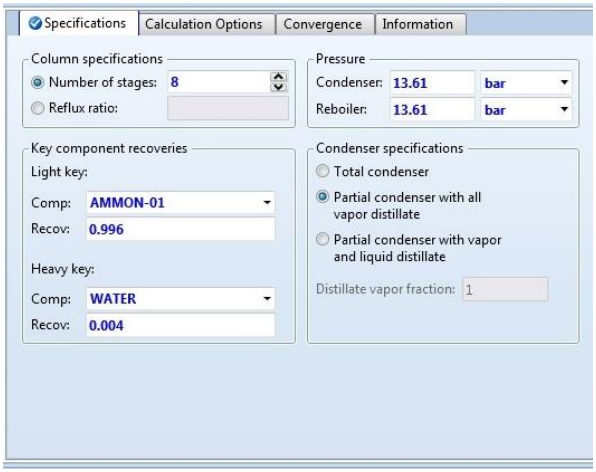


Fig- 3.4 Inputs for separator

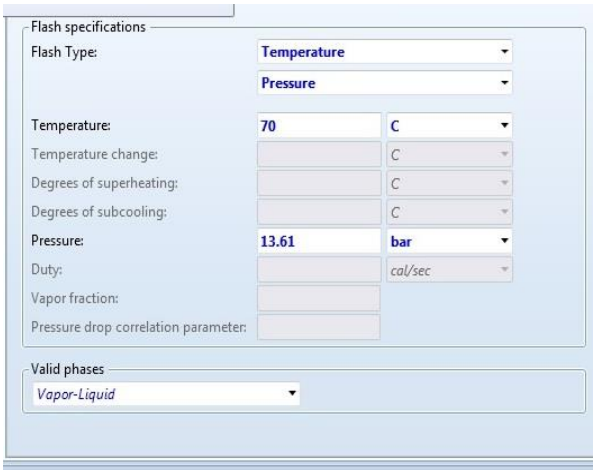


Fig- 3.5 Inputs for generator

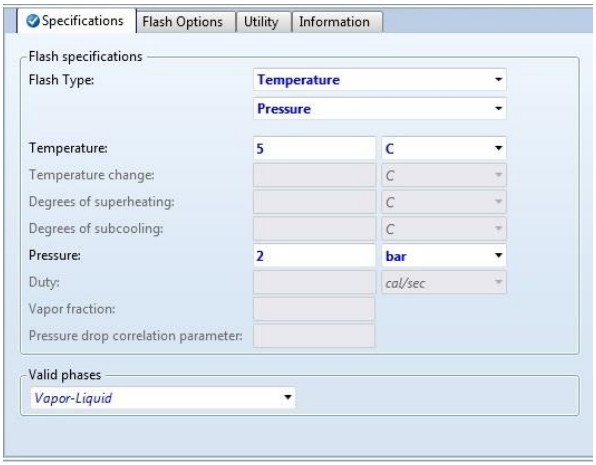


Fig- 3.6 Inputs for evaporator

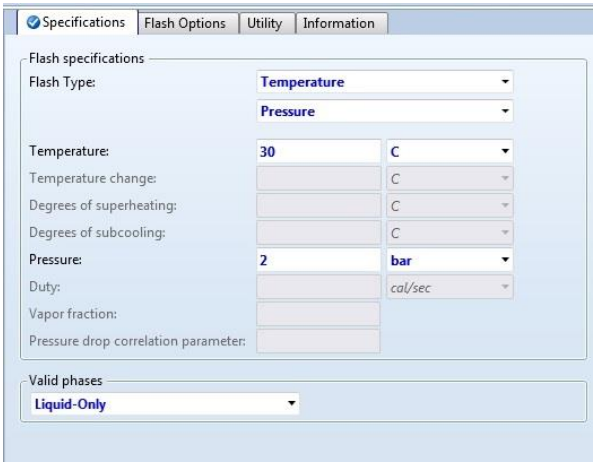


Fig- 3.7 Inputs for absorber

3.8 OUTPUTS FOR VARIOUS BLOCKS

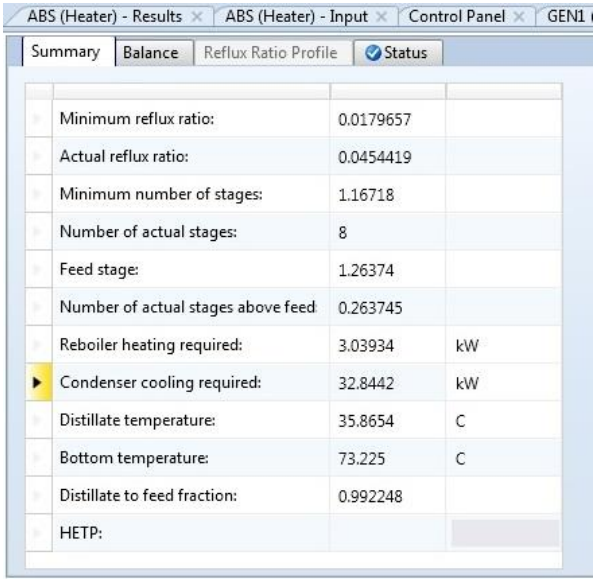


Fig- 3.8 Separator results

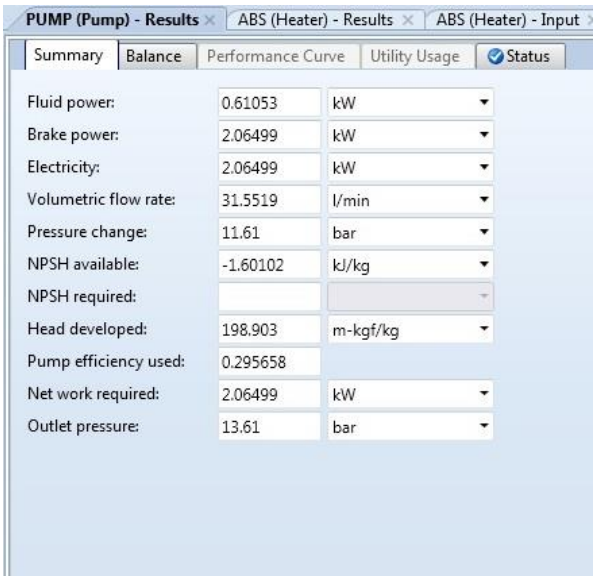


Fig- 3.9 Pump results

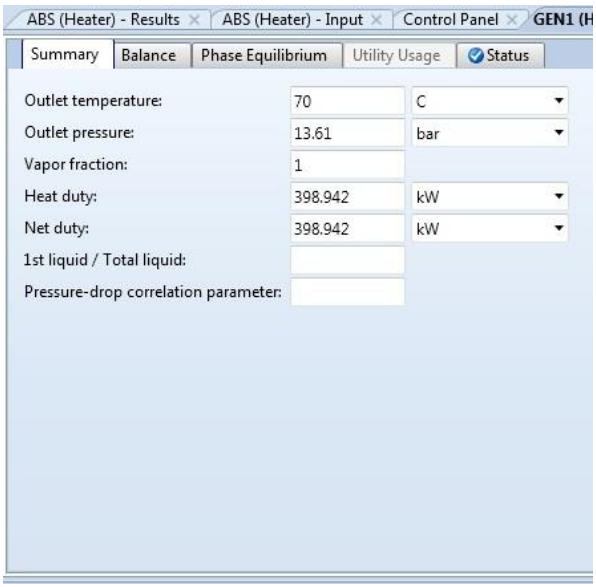


Fig- 3.10 Generator results

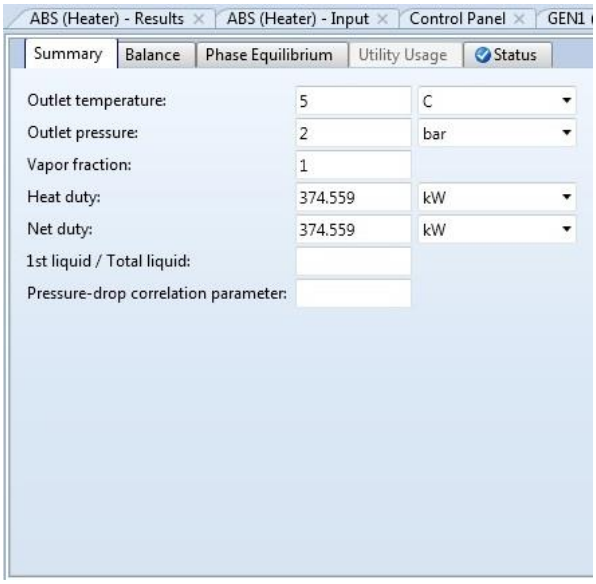


Fig- 3.11 Evaporator results

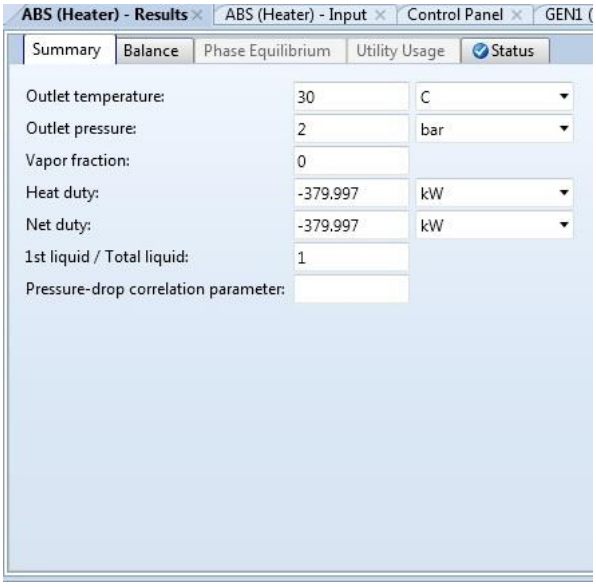


Fig- 3.12 Absorber results

Simulation of the cycle for different cycle parameters were carried out and the results for generator, evaporator, no. of separators etc. are tabulated in the following sections.

3.9 BLOCK RESULT FOR VARIOUS EVAPORATORTEMPERATURE (7 STAGE SEPARATOR) -

3.8.1 Generator Temperature = 70ºC

Table 3.3- Evaporator results

Evaporator Temperature (ºC)	Pressure (bar)	Vapour fraction	Heat Duty (kW)
5	2	1	373.87
6	2	1	374.52
7	2	1	375.17
8	2	1	375.82
10	2	1	374.56

Table 3.4- Generator results

Evaporator Temperature (ºC)	Pressure (bar)	Vapour fraction	Heat Duty (kW)
5	13.61	1	398.9
6	13.61	1	398.9
7	13.61	1	398.9
8	13.61	1	398.9
10	13.61	1	398.9

Table 3.5- Separator results

Evaporator Temp.(ºC)	Minimum RR	Actual RR	Minimum no. of stages	Actual no. of stages	Feed stage	Separator Duty (kW)	Distillate Temp. (ºC)	Bottom temperature (ºC)
5	0.018	0.05	1.17	7	1.23	42.17	35.86	73.23
6	0.018	0.05	1.17	7	1.23	42.17	35.86	73.23
7	0.018	0.05	1.17	7	1.23	42.17	35.86	73.23
8	0.018	0.05	1.17	7	1.23	42.17	35.86	73.23
10	0.018	0.05	1.17	7	1.23	42.17	35.86	73.23

3.8.2 Generator Temperature = 75ºC

Table 3.6- Evaporator results-

Evaporator Temperature (ºC)	Pressure (bar)	Vapour fraction	Heat Duty (kW)
5	2	1	374.56
6	2	1	374.56
7	2	1	374.56
8	2	1	374.56
10	2	1	374.56

Table 3.7- Generator results-

Evaporator Temperature (ºC)	Pressure (bar)	Vapour fraction	Heat Duty (kW)
5	13.61	1	402.65
6	13.61	1	402.65
7	13.61	1	402.65
8	13.61	1	402.65
10	13.61	1	402.65

Table 3.8- Separator results-

Evaporator Temp.(ºC)	Minimum RR	Actual RR	Minimum no. of stages	Actual no. of stages	Feed stage	Separator Duty (kW)	Distillate Temp. (ºC)	Bottom temperature (ºC)
5	0.028	0.06	1.17	7	1.23	49.27	35.86	73.23
6	0.028	0.06	1.17	7	1.23	49.27	35.86	73.23
7	0.028	0.06	1.17	7	1.23	49.27	35.86	73.23
8	0.028	0.06	1.17	7	1.23	49.27	35.86	73.23
10	0.028	0.06	1.17	7	1.23	49.27	35.86	73.23

3.8.3 Generator Temperature = 80ºC

Table 3.9- Evaporator result

Evaporator Temperature (ºC)	Pressure (bar)	Vapour fraction	Heat Duty (kW)
5	2	1	374.56
6	2	1	374.56
7	2	1	374.56
8	2	1	374.56
10	2	1	374.56

Table 3.10- Generator result

Evaporator Temperature (ºC)	Pressure (bar)	Vapour fraction	Heat Duty (kW)
5	13.61	1	406.36
6	13.61	1	406.36
7	13.61	1	406.36
8	13.61	1	406.36
10	13.61	1	406.36

Table 3.11- Separator results

Evaporator Temp.(ºC)	Minimum RR	Actual RR	Minimum no. of stages	Actual no. of stages	Feed stage	Separator Duty (kW)	Distillate Temp. (ºC)	Bottom temperature (ºC)
5	0.037	0.07	1.17	7	1.23	56.5	35.86	73.23
6	0.037	0.07	1.17	7	1.23	56.5	35.86	73.23
7	0.037	0.07	1.17	7	1.23	56.5	35.86	73.23
8	0.037	0.07	1.17	7	1.23	56.5	35.86	73.23
10	0.037	0.07	1.17	7	1.23	56.5	35.86	73.23

From Table 3.3 to Table 3.10 we observe various block results for different evaporator temperatures.

3.10 BLOCK RESULT FOR VARIOUS GENERATOR TEMPERATURE (8 STAGE SEPARATOR) -

Evaporator Temperature = 5 ºC

Table 3.12- Evaporator results

Generator Temperature (ºC)	Pressure (bar)	Vapour fraction	Heat Duty (kW)
70	2	1	374.56
75	2	1	374.56
80	2	1	374.56
85	2	1	374.56
90	2	1	374.56

Table 3.13- Generator results

Generator Temperature (ºC)	Pressure (bar)	Vapour fraction	Heat Duty (kW)
70	13.61	1	398.94
75	13.61	1	402.65
80	13.61	1	406.36
85	13.61	1	410.08
90	13.61	1	413.81

Table 3.14- Separator results

Generator Temp.(ºC)	Minimum RR	Actual RR	Minimum no. of stages	Actual no. of stages	Feed stage	Separator Duty (kW)	Distillate Temp. (ºC)	Bottom temperature (ºC)
70	0.018	0.045	1.17	8	1.26	35.87	35.86	73.23
75	0.027	0.055	1.17	8	1.26	42.96	35.86	73.23
80	0.037	0.065	1.17	8	1.26	50.12	35.86	73.23
85	0.046	0.074	1.17	8	1.26	57.33	35.86	73.23
90	0.056	0.085	1.17	8	1.26	64.54	35.86	73.23

From Table 3.11 to Table 3.17 we observe various block results for different generator temperatures.

3.11 BLOCK RESULTS FOR VARIOUS NO. OF STAGES AT DIFFERENT GENERATOR TEMPERATURE (T_E = 5 ^{O}C) –

3.10.1 Generator Temperature = 70ºC

Table 3.15- Evaporator result

No. of Stages	Pressure (bar)	Vapour fraction	Heat Duty (kW)
3	2	1	374.56
5	2	1	374.56
7	2	1	374.56
9	2	1	374.56
15	2	1	374.56

Table 3.16- Generator results

No. of Stages	Pressure (bar)	Vapour fraction	Heat Duty (kW)
3	13.61	1	398.94
5	13.61	1	398.94
7	13.61	1	398.94
9	13.61	1	398.94
15	13.61	1	398.94

Table 3.17- Separator result

No. of Stages	Minimum RR	Actual RR	Minimum no. of stages	Actual no. of stages	Feed stage	Separator Duty (kW)	Distillate Temp. (ºC)	Bottom temperature (ºC)
3	0.018	0.18	1.17	3	1.1	200.26	35.86	73.23
5	0.018	0.07	1.17	5	1.17	65.56	35.86	73.23
7	0.018	0.05	1.17	7	1.23	42.14	35.86	73.23
9	0.018	0.04	1.17	9	1.3	31.3	35.86	73.23
15	0.018	0.03	1.17	15	1.5	29.81	35.86	73.23

3.10.2 Generator Temperature = 75ºC

Table 3.18- Evaporator result

No. of Stages	Pressure (bar)	Vapour fraction	Heat Duty (kW)
3	2	1	374.56
5	2	1	374.56
7	2	1	374.56
9	2	1	374.56
15	2	1	374.56

Table 3.19- Generator results

No. of Stages	Pressure (bar)	Vapour fraction	Heat Duty (kW)
3	13.61	1	402.65
5	13.61	1	402.65
7	13.61	1	402.65
9	13.61	1	402.65
15	13.61	1	402.65

Table 3.20- Separator result

No. of Stages	Minimum RR	Actual RR	Minimum no. of stages	Actual no. of stages	Feed stage	Separator Duty (kW)	Distillate Temp. (ºC)	Bottom temperature (ºC)
3	0.027	0.19	1.17	3	1.1	208.9	35.86	73.23
5	0.027	0.08	1.17	5	1.16	72.9	35.86	73.23
7	0.027	0.06	1.17	7	1.23	49.3	35.86	73.23
9	0.027	0.05	1.17	9	1.29	38.3	35.86	73.23
15	0.027	0.04	1.17	15	1.49	33.13	35.86	73.23

3.10.3 Generator Temperature = 80ºC

Table 3.21- Evaporator result

No. of Stages	Pressure (bar)	Vapour fraction	Heat Duty (kW)
3	2	1	374.56
5	2	1	374.56
7	2	1	374.56
9	2	1	374.56
15	2	1	374.56

Table 3.22- Generator results

No. of Stages	Pressure (bar)	Vapour fraction	Heat Duty (kW)
3	13.61	1	406.36
5	13.61	1	406.36
7	13.61	1	406.36
9	13.61	1	406.36
15	13.61	1	406.36

Table 3.23- Separator result

No. of Stages	Minimum RR	Actual RR	Minimum no. of stages	Actual no. of stages	Feed stage	Separator Duty (kW)	Distillate Temp. (ºC)	Bottom temperature (ºC)
3	0.037	0.21	1.17	3	1.1	213.6	35.86	73.23
5	0.037	0.09	1.17	5	1.16	80.35	35.86	73.23
7	0.037	0.07	1.17	7	1.23	56.49	35.86	73.23
9	0.037	0.06	1.17	9	1.29	45.23	35.86	73.23
15	0.037	0.05	1.17	15	1.49	37.73	35.86	73.23

From Table 3.15 to Table 3.23 we observe various block results for different no. of stages.

CHAPTER 4

OPTIMIZATION, DESIGN AND ECONOMIC FEASIBILITY OF ABSORPTION REFRIGERATION SYSTEM USING AMMONIA+WATER AS WORKING PAIR

Based on the results from all the simulation in chapter 3-

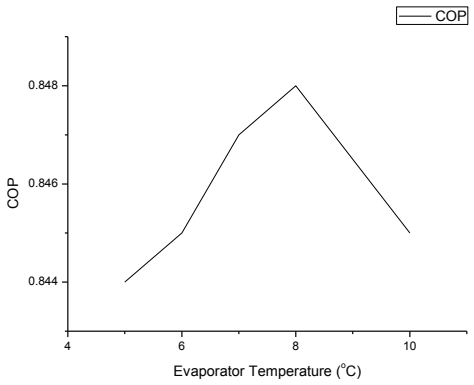
4.1 EFFECT OF EVAPORATOR TEMPERATURE ON COP-

No of stages in separator = 7

Generator temperature = 70 ^{o}C

Table 4.1-

Evaporator Temperature(oC)	Evaporator Duty(kW)	Generator Duty(kW)	Pump Work(kW)	COP
5	373.87	444.04	2.065	0.844
6	374.52	444.04	2.065	0.845
7	375.17	444.04	2.065	0.847
8	375.82	444.04	2.065	0.848
10	374.56	444.04	2.065	0.845

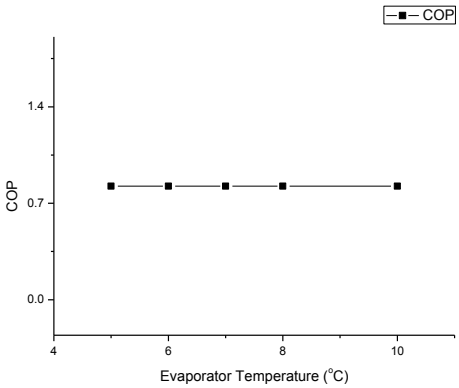


From the Table 4.1 we infer that with increase in evaporator temperature the COP values changes but quite insignificantly, it varies between 0.844 and 0.848.

Generator temperature = 75 ^{o}C

Table 4.2-

Evaporator Temperature(oC)	Evaporator Duty(kW)	Generator Duty(kW)	Pump Work(kW)	COP
5	374.56	451.93	2.065	0.825
6	374.56	451.93	2.065	0.825
7	374.56	451.93	2.065	0.825
8	374.56	451.93	2.065	0.825
10	374.56	451.93	2.065	0.825

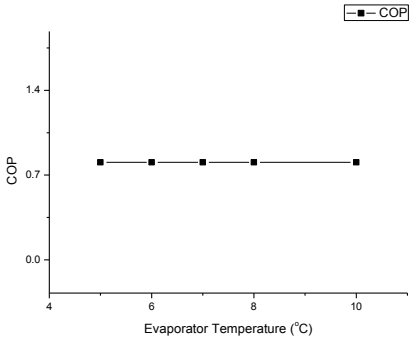


From Table 4.2 we infer that there is no change in COP with increase in evaporator temperature.

Generator temperature = 80 ^{o}C

Table 4.3-

Evaporator Temperature(oC)	Evaporator Duty(kW)	Generator Duty(kW)	Pump Work(kW)	COP
5	374.56	406.364	2.065	0.806
6	374.56	406.364	2.065	0.806
7	374.56	406.364	2.065	0.806
8	374.56	406.364	2.065	0.806
10	374.56	406.364	2.065	0.806



From Table 4.3 we infer that there is no change in COP of the system with increase in evaporator temperature.

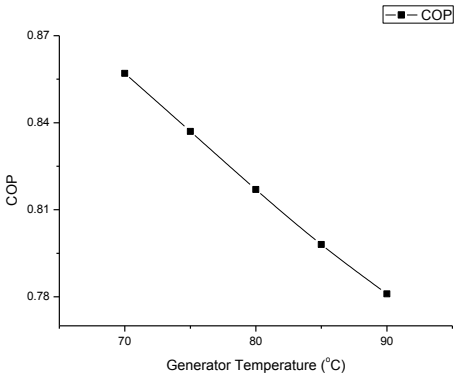
4.2 EFFECT OF GENERATOR TEMPERATURE ON COP-

No. of stages in separator = 8

Evaporator temperature = 5 ^{o}C

Table 4.4-

Generator Temperature(oC)	Evaporator Duty(kW)	Generator Duty(kW)	Pump Work(kW)	COP
70	374.56	434.82	2.065	0.857
75	374.56	445.61	2.065	0.837
80	374.56	456.48	2.065	0.817
85	374.56	467.43	2.065	0.798
90	374.56	478.32	2.065	0.781



From Table 4.4 we infer that with increase in the generator temperature the value of COP goes on decreasing and decreases by appreciable amount.

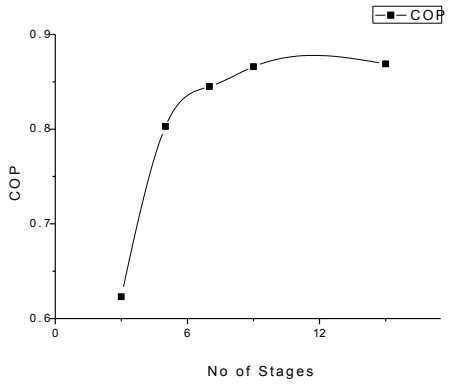
4.3 EFFECT OF NO. OF STAGES IN SEPARATOR ON COP-

Evaporator Temperature = 5 ^{o}C

1. Generator Temperature = 70 ^{o}C

Table 4.5-

No. of Stages	Evaporator Duty(kW)	Generator Duty(kW)	Pump Work(kW)	COP
3	374.56	598.94	2.065	0.623
5	374.56	464.51	2.065	0.803
7	374.56	441.08	2.065	0.845
9	374.56	430.24	2.065	0.866
15	374.56	428.75	2.065	0.869

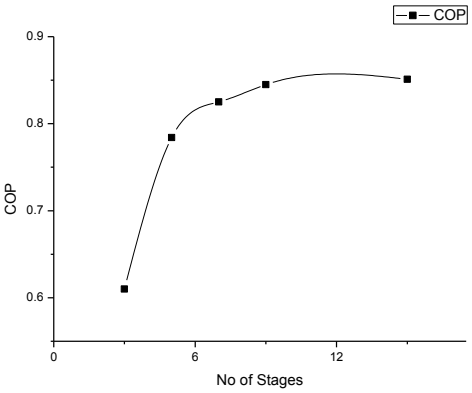


From Table 4.5 we infer that the COP value for the system increases appreciably with increase in no. of stages in separator, but after 9 stages the COP does not change appreciably with increase in no. of stages.

2. Generator Temperature = 75 ^{o}C

Table 4.6-

No. of Stages	Evaporator Duty(kW)	Generator Duty(kW)	Pump Work(kW)	COP
3	374.56	611.55	2.065	0.611
5	374.56	475.55	2.065	0.784
7	374.56	451.95	2.065	0.825
9	374.56	440.95	2.065	0.845
15	374.56	436.18	2.065	0.851

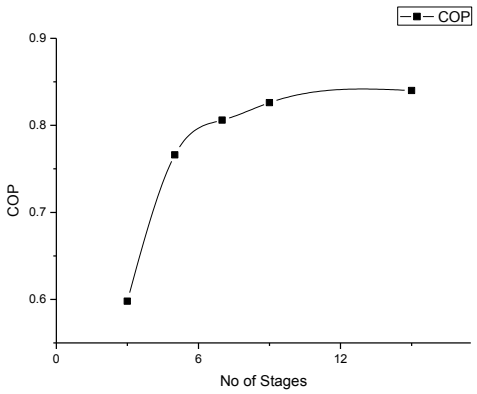


From Table 4.6 we infer that the COP value for the system increases appreciably with increase in no. of stages in separator, but after 9 stages the COP does not change appreciably with increase in no. of stages. As compared to T_G = 70ºC, the COP values for T_G = 75ºC are much less.

Generator Temperature = 80 ^{o}C

Table 4.7-

No. of Stages	Evaporator Duty(kW)	Generator Duty(kW)	Pump Work(kW)	COP
3	374.56	623.98	2.065	0.598
5	374.56	486.71	2.065	0.766
7	374.56	462.85	2.065	0.806
9	374.56	451.53	2.065	0.826
15	374.56	443.59	2.065	0.838



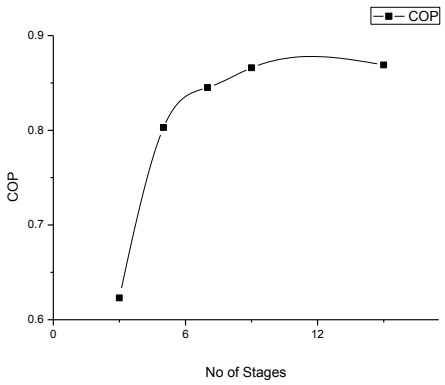
From Table 4.7 we infer that the COP value for the system increases appreciably with increase in no. of stages in separator, but after 9 stages the COP does not change appreciably with increase in no. of stages. As compared to T_G = 75^{o}C, the COP values for T_G = 80^{o}C are much less.

Evaporator Temperature = 7 ^{o}C

Generator Temperature = 70 ^{o}C

Table 4.8-

No. of Stages	Evaporator Duty(kW)	Generator Duty(kW)	Pump Work(kW)	COP
3	374.56	598.94	2.065	0.623
5	374.56	464.51	2.065	0.803
7	374.56	441.08	2.065	0.845
9	374.56	430.24	2.065	0.866
15	374.56	428.75	2.065	0.869



From Table 4.8 we infer that the COP value for the system increases appreciably with increase in no. of stages in separator, but after 9 stages the COP does not change appreciably with increase in no. of stages. We infer from the results in Table 4.8 that COP values does not change if we increase the evaporator temperature.

Since the effect on COP by varying evaporator temperature is quite insignificant, the analysis for 1 evaporator temperature is done. From the above analysis we conclude that the optimum conditions for operation of the absorption refrigeration cycle are-

1. Generator Temperature= 70 ^{o}C
2. Evaporator Temperature= 5 ^{o}C
3. Condenser Temperature= 33 ^{o}C
4. Absorber Temperature= 30 ^{o}C
5. No of Stages= 8

4.4 FEASIBLITY STUDY-

Basis at optimum conditions-

Plant operating for 10 hours/day and 300 days annually.

Electricity cost- $\mathbf{Rs}\frac{\mathbf{6.90}}{\mathbf{kWh}}$

Fixed cost for ammonia + water system = \$7000 = $\text{Rs}\frac{420000}{10\text{ years}}$

Per year fixed cost = $\mathbf{Rs\ 42000}$

Operating cost at optimum condition = $\text{Heat duty} * \text{Electricity cost}$

Heat Duty for 100 TR refrigeration-

1. Evaporator = $374.56 * 10 * 300 = 1123680$ kWh
2. Generator = $400 * 10 * 300 = 1200000$ kWh
3. Condenser = $376.76 * 10 * 300 = 1103280$ kWh
4. Pump Work = $2.065 * 10 * 300 = 6195$ kWh
5. Absorber = $380 * 10 * 300 = 1140000$ kWh

Total heat duty per year = $Q_E + Q_G + W_P = \mathbf{2329875\ kWh}$

For 1 TR refrigeration operating cost = $23298.75 * 6.90 = \text{Rs}\ \mathbf{160761.375}$

Cost of 99.6 % ammonia = $\frac{\mathbf{\$600}}{\mathbf{ton}}$

Comparison of ARS with VCR –

1. VCR
   Fixed cost of a 1 ton VCR = $\mathbf{Rs\ 40000}$

   Lifetime of VCR = 4 years

   So average fixed cost of VCR = Rs 20000

   Heat duty of a 1 ton VCR = 1.5 kW

   Annual heat duty = $1.5 * 10 * 300 = 4500$ kW

   Annual cost operating cost of VCR = $4500 * 6.90 = \text{Rs}\ 31050$

   Total annual cost of VCR = $\text{Rs}\ 31050 + \text{Rs}\ 20000 = \text{Rs}\ 51050$

2. ARS

Total cost of operation = Fixed cost + pump work = $42000 + 427.5$ $= \mathbf{Rs\ 42427.5}$

Thus profit of using ARS = Total cost of VCR − Total cost of ARS $= 51050 - 42427.5$

=$\mathbf{Rs\ 8622.5}$

So annual savings by using ARS is $\mathbf{Rs\ 8622.5}$

From the feasibility analysis we observe that for a refrigeration of 1 TR, we require 23 MWh low grade heat at a temperature of 70ºC. Further, if we change the refrigeration load the requirement of low grade heat also changes.

CHAPTER 5

CONCLUSIONS AND FUTURE RECOMMENDATIONS

5.1 CONCLUSIONS-

The absorption refrigeration system chosen here is an established system. In spite of this being well versed we have taken up this system to pave out ways for future study of ARS and obtain the optimizing parameter for this cycle. We have ammonia + water as the working pair and operated the cycle for various temperature ranges of the evaporator, generator and absorber, we also worked on different no. of stages to check the COP of the system. In the process we obtain the optimum parameters for smooth operation of the cycle. Based on the optimum results we made a thermo-economic analysis and made the feasibility study of the cycle. The result obtained is quite encouraging as it shows that using an ARS system instead of VCR the plant runs in a net profit.

The evaporator temperature has a very insignificant effect on the COP of the system and thus does not change the COP even if we increase the temperature. The generator has a very vivid impact on the COP. As we increase the generator temperature the COP value decreases. Thus further concreting our base for using low grade heat as the source of power for operating the generator. Further, no. of stages of the separator also effect the COP of the system. Increasing the no. of stages increases the COP up to a certain level but on further increasing the stages the COP becomes constant.

Thus the optimum conditions comes as- T_G = 70 ºC, T_E = 5 ºC, T_C = 33 ºC, T_A = 30 ºC and the optimum no. of stages for the separator comes to 8.

The feasibility study also suggests that the use same capacity (1 TR) ARS instead of VCR is profitable by an amount of INR 8.5K annually. From the feasibility analysis we observe that for a refrigeration of 1 TR, we require 23 MWh low grade heat at a temperature of 70ºC. Further, if we change the refrigeration load the requirement of low grade heat also changes.

5.2 FUTURE RECOMMENDATIONS-

The optimization and design used in this analysis is not very rigorous, further study can be done on component basis for each blocks used and then the thermos-economic feasibility of the system can be calculated. Ammonia+ water system is a very old ARS, further study can be done on these systems using different working pairs, especially the ionic liquid pairs. Further research can be done on using the best possible combination of blocks and an elaborate study on the optimization aspect using the flowrate analysis, the refrigerant concentration analysis and the heat exchanger design can also be done. There was a severe problem on using a valid separator for regenerating the refrigerant occurred and a DSTWU column was used for the same. Analysis can be carried out on the use of various separators for the purpose of regeneration. The system used is a high pressure system which has high energy demand, so study must be carried out in choosing a more suitable working pair for use in the absorption refrigeration system.

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